mented (WSRC 1998e) and forms the basis for the environmental impact analysis presented in this SEIS.

A.4.2 RECOMMENDATION AND REVIEW

On October 29, 1998, following review by the WSRC Review Panel Team, WSRC recommended to DOE the Small Tank Precipitation process as the most reasonable replacement salt processing technology and the Ion Exchange process as a backup technology (WSRC 1998f).

A DOE Savannah River (SR) Review Team evaluated the WSRC recommendation and concluded that the remaining technical uncertainties for each of the alternatives were too significant to justify selection of a preferred technology (DOE 1998b). The DOE-SR Review Team recommended that additional research and development be conducted to address the key technical uncertainties associated with the two technologies, so that one could be identified as most reasonable. The Review Team agreed with WSRC that one of the four technologies considered in Phase III, Solvent Extraction, should be eliminated from further consideration because of its insufficient technical maturity. The DOE Review Team concluded that the Direct Disposal in Grout alternative should not be eliminated, based on its potential to reduce construction and operating costs and the high confidence in its technology, safety, and feasibility for implementation.

A DOE-Headquarters Independent Review Team concluded that both the Small Tank Precipitation and the Ion Exchange alternatives were technically feasible. This team agreed with the SET that Direct Disposal in Grout should be eliminated from further consideration, because of regulatory issues that had the potential to substantially increase the time required to implement the technology (DOE 1998c). DOE concluded that further investigations of this alternative would not be pursued as long as a cesium-separation technology could be proved technically and economically practical.

In January 1999, DOE directed WSRC to conduct additional research and development on the Small Tank Precipitation and Ion Exchange alternatives. These additional studies concluded with WSRC maintaining its recommendation to pursue design and construction for the Small Tank Precipitation process (WRSC 1999b,c). WSRC further noted that, with additional development to reduce technical and engineering risk factors, the Ion Exchange process could also prove suitable for SRS, as well as a DOE complex-wide application for salt processing. TC

During this period, the technology for the Solvent Extraction process advanced independent of the SRS alternative evaluations. This information, coupled with recommendations from the National Research Council of the National Academy of Sciences (NAS 1999, 2000) and improved understanding of tank farm water management issues (WSRC 1999a), led the Department to reconsider the potential to mature and implement this alternative in time to support HLW salt processing needs.

In February 2000, DOE requested WSRC to initiate further development of the Solvent Extraction alternative, aimed at the timely resolution of previously identified problems (DOE 2000). Consequently, the Solvent Extraction technology is included as a reasonable alternative in the SEIS.

A.4.3 PROCESS DESCRIPTIONS

A.4.3.1 Small Tank Tetraphenylborate Precipitation

In the Small Tank Precipitation technology (WSRC 1998e,g,h), the soluble salt components of the wastes would be processed using precipitation-sorption procedures analogous to the ITP process to separate cesium and other soluble constituents from the waste solutions. The process would be conducted as a continuous operation in stirred small tanks (15,000 gallons) with the solution agitated constantly to avoid excessive decomposition of tetraphenylborate and accompanying generation of benzene before separation. In the Small Tank Precipitation technology, tetraphenylborate solution would be added to precipitate cesium and potassium, and a slurry of monosodium titanate would be added to sorb residual strontium and actinides from the salt solution. The resulting solids, along with residual sludge, would be concentrated by filtration and washed to remove soluble salts, then treated chemically to convert the precipitate to a non-flammable form for transfer to DWPF. Catalytic decomposition of the precipitate, with removal of the benzene formed, would generate a product stream containing cesium in aqueous solution and strontium and actinides sorbed onto monosodium titanate for vitrification. The low activity salt solution recovered by filtration would be transferred to the Saltstone Manufacturing and Disposal Facility for processing. The wash water would be recycled into the incoming soluble salt solution.

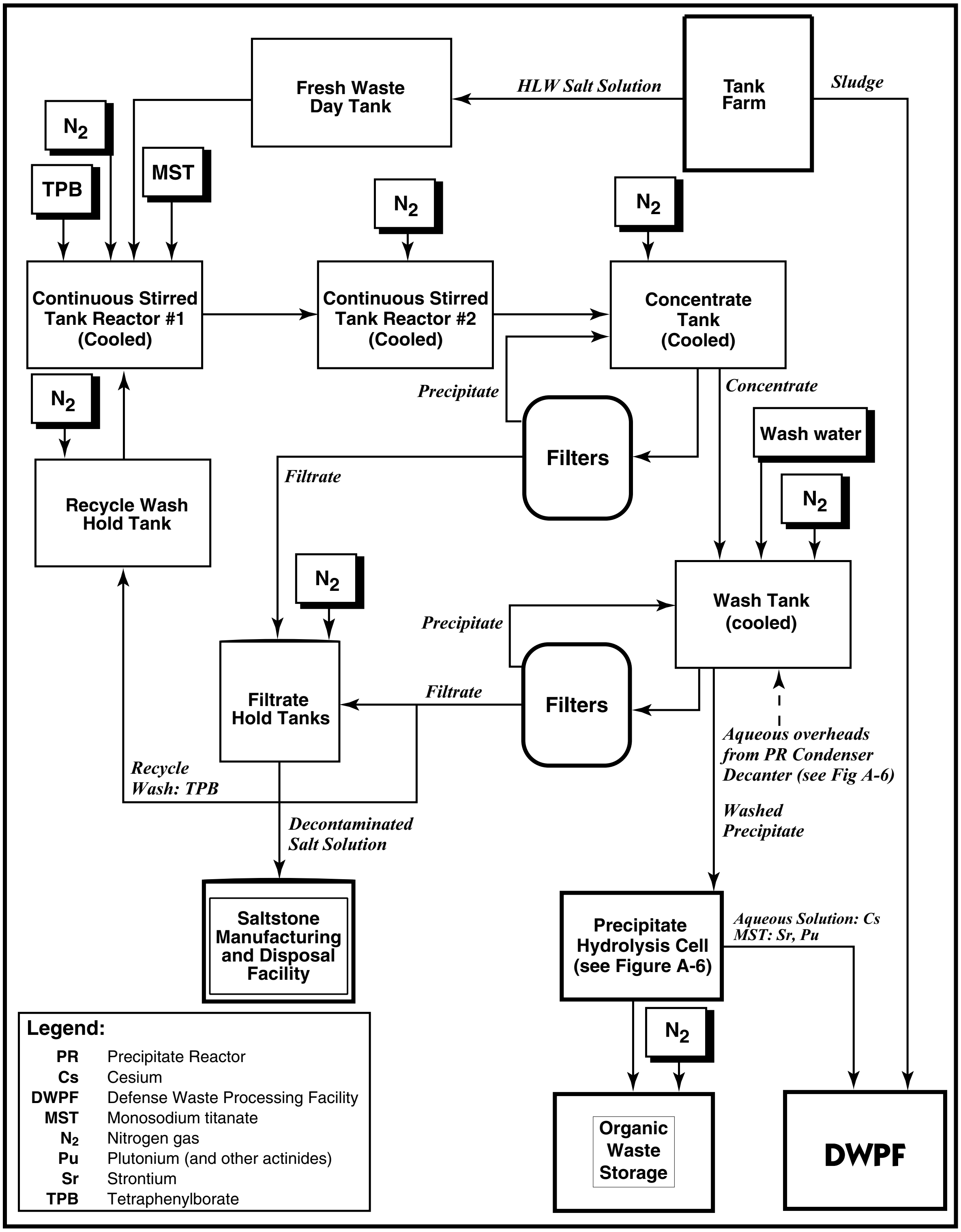
Small Tank Precipitation would be performed in a new facility to be constructed at Site B in S Area. Process flows for the Small Tank Precipitation alternative are shown in Figure A-5. Salt solution would be collected in an H-Area tank and pumped to the Small Tank Precipitation facility. A section of new interarea transfer line would be required to connect the new facility to the existing transfer line. The precipitation process would be conducted in two Continuous Stirred Tank Reactors. Salt solution mixed with tetraphenylborate, monosodium titanate, process water, and recycled wash water in the first tank reactor would flow to the second tank reactor, providing reaction conditions needed to maximize decontamination factors for the precipitation and sorption processes.

The precipitate slurry, containing about one weight percent tetraphenylborate and monosodium titanate solids, would be transferred continuously from the second tank reactor to a Concentrate Tank, where it would be concentrated to about 10 weight percent solids by cross-flow filtration. The resulting filtrate would be pumped to a Filtrate Hold Tank for later transfer to the Saltstone Manufacturing and Disposal Facility for immobilization in the saltstone vaults.

The precipitate slurry accumulated in the Concentrate Tank would be transferred to the Wash Tank for washing in a batch process to remove soluble sodium salts. Spent wash water would be separated from the precipitate by cross-flow filtration. The washed precipitate would be treated in the Precipitate Hydrolysis Cell (PHC) of the facility to eliminate benzene and generate an aqueous product stream termed Precipitate Hydrolysis Aqueous (PHA).

The PHC incorporates process operations formerly assigned to the Salt Processing Cell of DWPF (see text box below). Process flows for the PHC are shown in Figure A-6. In the PHC, the washed precipitate would be combined with a copper nitrate-formic acid solution in the Precipitate Reactor to catalytically decompose the tetraphenylborate precipitate. The Precipitate Reactor would be heated to boiling and the benzene would be removed as it was formed. The benzene and water vaporized during boiling would be condensed in the Precipitate Reactor Condenser, with aqueous and organic condensates separated by decantation for return to the Precipitate Evaporator and Organic Evaporator, respectively. After a period of reflux boiling, the PHA product would be concentrated by distillation, with the aqueous overheads transferred to the Precipitate Wash Tank.

A second evaporation would be conducted in the PHC to ensure that the separated organic was sufficiently decontaminated for transfer outside the containment area. Wash water would be added to the Organic Evaporator and the boiling, evaporation, and decantation cycle would be repeated, with the twice-distilled benzene collected in the Organic Evaporator Condensate Tank for transfer to the Organic Waste Storage Tank.



NW SDA EIS/Grfx/A-5 Sm TPB.ai

Figure A-5. Small Tank Precipitation process flow diagram.

Transfer of DWPF Salt Processing Cell Operations to Small Tank Precipitation Facility

The design basis for the Small Tank Precipitation facility was modified to include the precipitate decomposition operations previously programmed for the DWPF. These operations, to be conducted in a Precipitate Hydrolysis Cell (PHC), had been tested during DWPF nonradioactive process runs, but were not employed during radioactive (sludge only) processing because of the unavailability of ITP feed. Major justifications for transferring the PHC operations to the Small Tank Processing facility are as follows:

- Safety – Lessons learned in DWPF design would provide PHC equipment with increased safety and control margins. As redesigned, the equipment would operate under slight positive pressure and low purge rates of inert cover gas.
- Capacity – Increased throughputs of PHC equipment would provide Small Tank Precipitation processing capacity needed to match required HLW salt removal schedules, with a substantial reduction in life-cycle processing time and significant cost savings.
- Flexibility – The vacated cell in the DWPF would become available for other potentially needed operations, including evaporation of DWPF recycle waste streams to conserve Tank Farm space pending startup of salt processing operations.
- Organic Disposition – Precipitate Hydrolysis Cell operations in the Small Tank Precipitation facility would confine generation and disposal of flammable organic byproducts to the process facility. This would avoid buildups of high-boiling organics in DWPF process and ventilation systems, and transfer in DWPF recycle streams to the Tank Farm. Lag storage and transfer to DWPF would be provided for the non-flammable aqueous product of the PHC operations, rather than the flammable tetraphenylborate precipitate product.

The tetraphenylborate employed in the Small Tank Precipitation process could undergo radiolytic and, under certain conditions, catalytic degradation, producing benzene before the decomposition reactions prescribed in the PHC. The Small Tank Precipitation process would require controlled benzene removal in all steps. Benzene production in the precipitation and washing operations would be limited by the continuous processing of relatively small waste volumes, by a short processing time, and by chilling the process vessels. Accumulation of benzene would be avoided by continuous agitation to prevent retention in the process mixtures and a flowing nitrogen gas blanket to sweep benzene vapors from the system. Benzene formation during precipitate decomposition in the PHC would be controlled by process constraints, with all process vessels purged with nitrogen to maintain oxygen concentrations below combustion limits.

A.4.3.2 Crystalline Silicotitanate Ion Exchange

The Ion Exchange Process (WSRC 1998e,i,j) would employ a crystalline silicotitanate particulate solid (resin) to remove the cesium from the salt solution. In the ion exchange reaction, the radioactive cesium displaces nonradioactive constituents (sodium) of the resin. As in the Small Tank Precipitation process, residual strontium and actinides in the salt solution would be sorbed onto monosodium titanate and, in conjunction with residual sludge, filtered from the salt solution prior to the crystalline silicotitanate ion exchange treatment. The cesium-loaded crystalline silicotitanate resin and the monosodium titanate solids would be transferred to DWPF as slurries to be combined with sludge for incorporation into the glass waste form. Low activity salt solution would be immobilized as saltstone in onsite vaults at the Saltstone Manufacturing and Disposal Facility.

The Ion Exchange process would be performed in a new facility built at Site B in S Area. Process operations are illustrated in the flow diagram in Figure A-7. Salt solution would be pumped from an H-Area tank to the Ion Exchange facility. A new feed line between the existing inter-area transfer line and the Ion Exchange facility would be required for this transfer. In initial feed clarification operations in the batch Alpha Sorption Tank, the salt solution would be mixed with monosodium titanate to sorb soluble strontium and actinides and then filtered by cross-

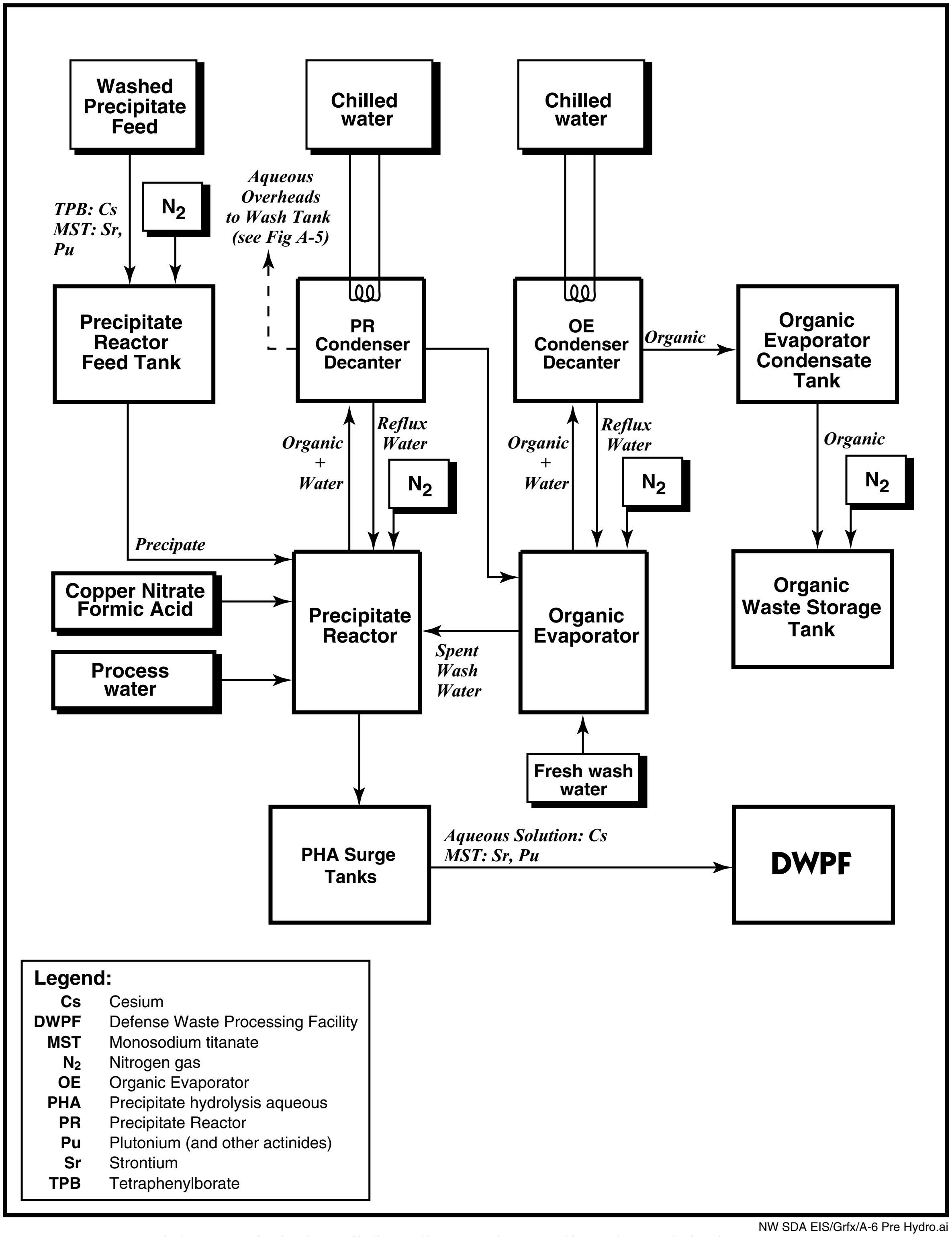
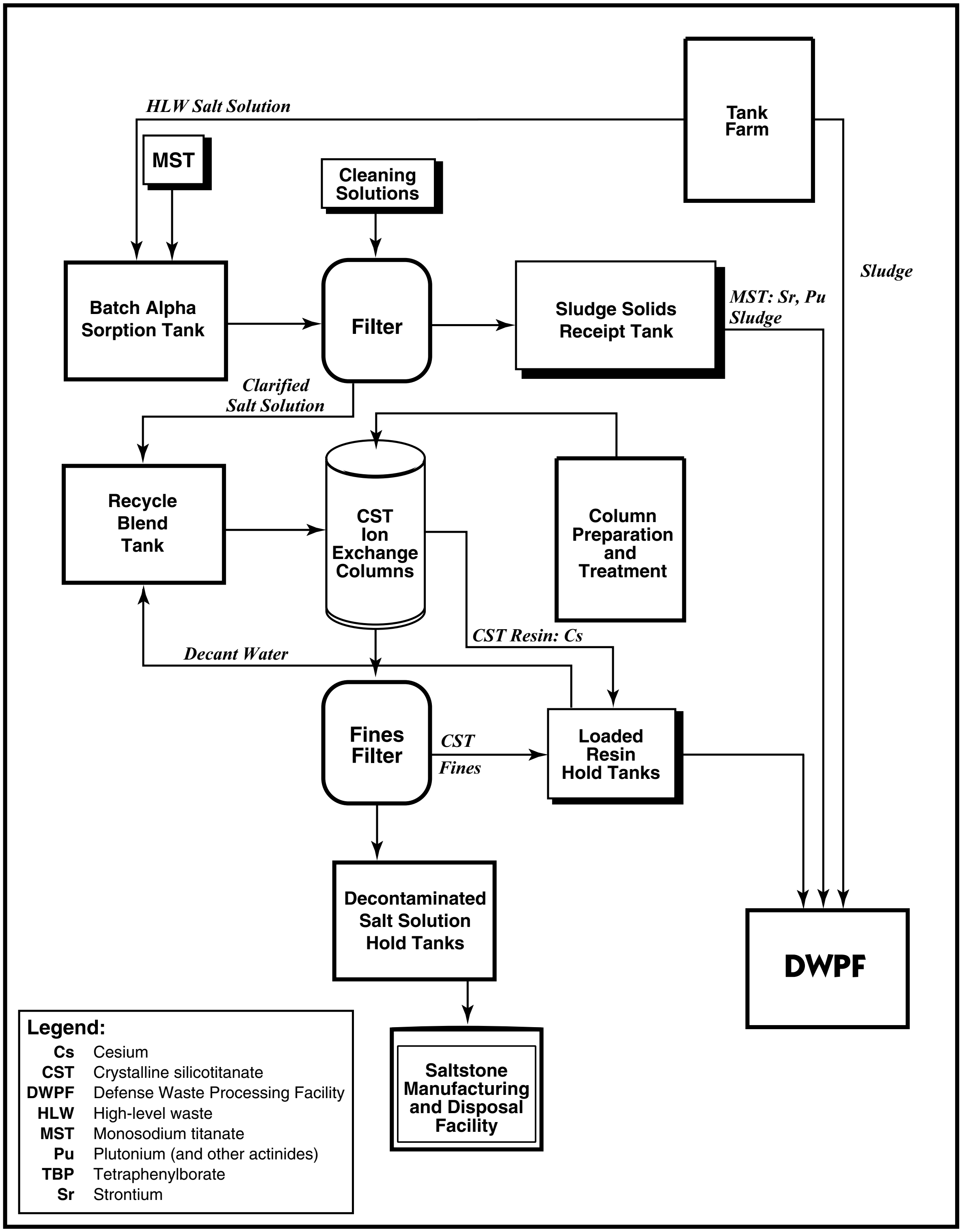


Figure A-6. Precipitate Hydrolysis Cell flow diagram for Small Tank Precipitation process.



NW SDA EIS/Grfx/A-7 CST Ion Flow.ai

Figure A-7. Ion Exchange process flow diagram.

flow filtration to remove monosodium titanate solids and residual sludge. These clarification operations would be necessary to prevent plugging of the ion exchange columns during subsequent processing of the salt solution. The product slurry, washed and concentrated to about 5 weight percent solids, would be pumped through new and existing transfer lines to DWPF as feed for the vitrification process.

After filtration, the clarified salt solution would be transferred to the Recycle Blend Tank in the Ion Exchange facility for dilution with process water, and pumped through a series of four ion exchange columns to remove radioactive cesium. Cesium transfer from the salt solution would take place in the first three columns, with the fourth column in reserve for use when the first column in the series reached saturation (> 90 percent maximum capacity) and was taken out of service. Saturated resin in the column would be flushed with water and pumped as slurry to DWPF. The first ion exchange column would then be replenished with fresh resin and held in reserve (as the fourth column) while cesium ion exchange took place in what had been the second, third, and fourth columns. The cycle would continue with the lead column reaching saturation and the reserve column becoming the last in the series of three operating columns. Low activity salt solution recovered as effluent from the third column would be filtered to prevent any cesium-loaded fine particles from recontaminating the salt solution. The low activity salt solution would be sampled in a Product Holdup Tank prior to transfer to the Decontaminated Salt Solution Hold Tanks, to ensure that requirements for disposal as saltstone were met. The low activity salt solution would be transferred to the Saltstone Manufacturing and Disposal Facility for disposal in onsite vaults. All process wastewater would be recycled and reused.

The Ion Exchange process would result in the accumulation of as much as 15 million curies of cesium within the processing cell. This radioactive loading would necessitate stringent shielding requirements and operational controls because of the generation of hydrogen and other gases.

A.4.3.3 Caustic Side Solvent Extraction

In the Solvent Extraction process (WSRC 1998e, 1999d), radioactive cesium would be separated from the caustic HLW salt solution by extracting it from the aqueous phase into an insoluble organic phase, thereby generating a low activity salt solution for immobilization in saltstone. The separated cesium, recovered from the organic phase by back extraction (stripping) into an acidic aqueous solution, would be transferred to DWPF for incorporation, along with HLW sludge, into the glass waste form. Prior processing of the HLW salt solution by monosodium titanate to remove soluble strontium and actinides, followed by filtration of monosodium titanate solids and residual sludge, would be necessary to meet saltstone acceptance limits and avoid interference of residual solids in the solvent extraction process.

The organic phase into which the cesium would be extracted is a kerosene-like solvent (diluent) containing an organic extractant (termed BoB-CalixC6) and a diluent modifier (typically Cs-7SBT). The extractant is highly specific for cesium, permitting separation from sodium by a factor of 10^4 (10,000) and from potassium by a factor of 10^2 (100). The diluent modifier increases the cesium extraction capability by increasing extractant solubility in the diluent. The subsequent stripping of separated cesium back into an aqueous solution is promoted by addition of a suppressor constituent, typically trioctylamine (TOA), to the organic phase. The TOA also mitigates the deleterious effects of impurities in the aqueous solution. Chemical structures and concentrations of the additions to the diluent organic phase are specified in the text box on page A-18.

The Solvent Extraction process would be performed in a new facility at Site B in S Area. Process operations are represented by the flow diagram in Figure A-8. In operations similar to

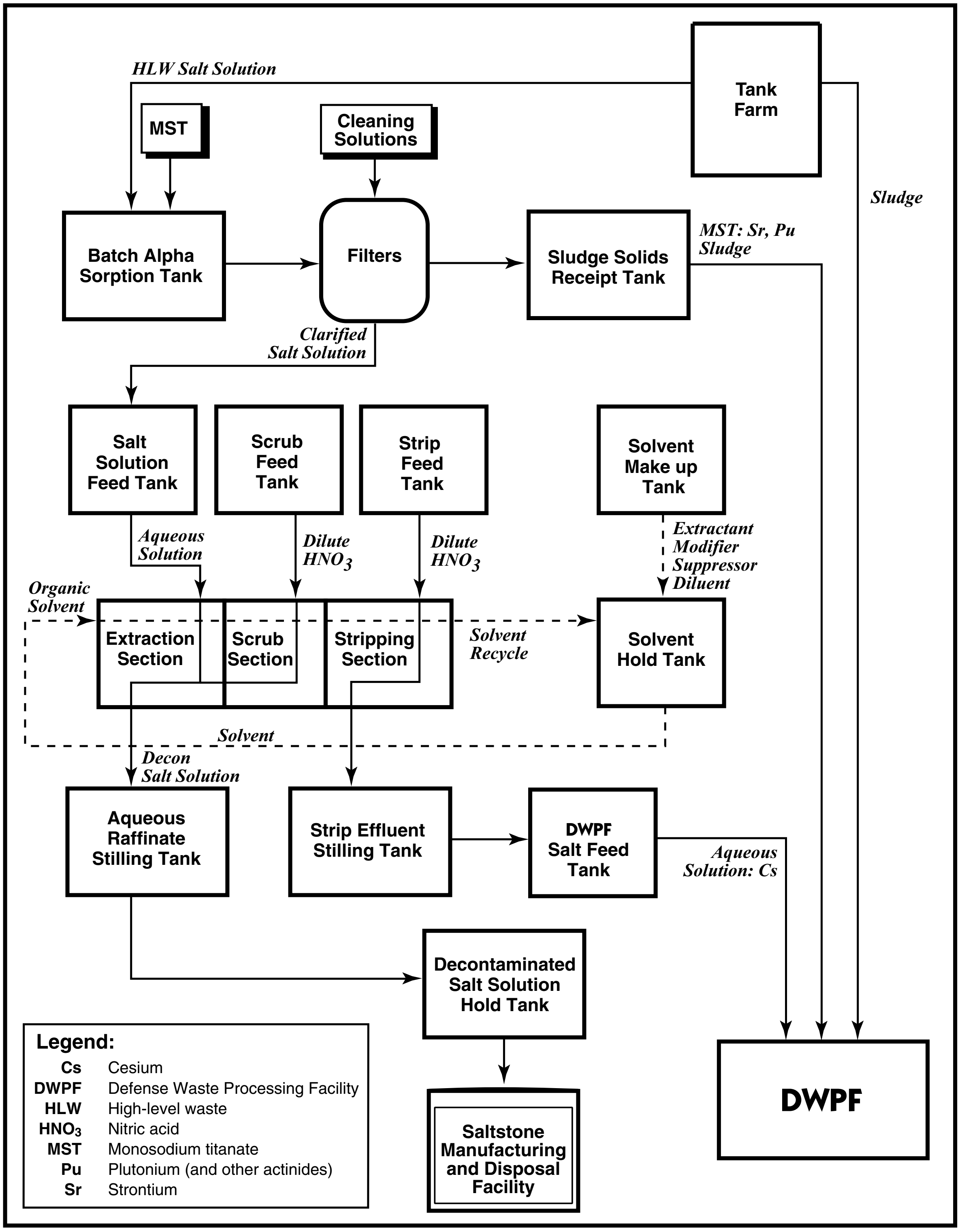
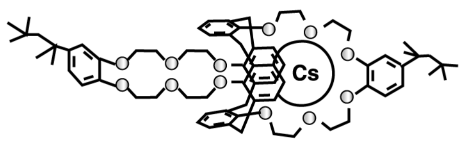
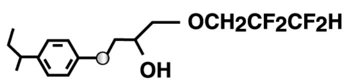


Figure A-8. Solvent Extraction process flow diagram.

Components of Organic Phase used in Solvent Extraction Process

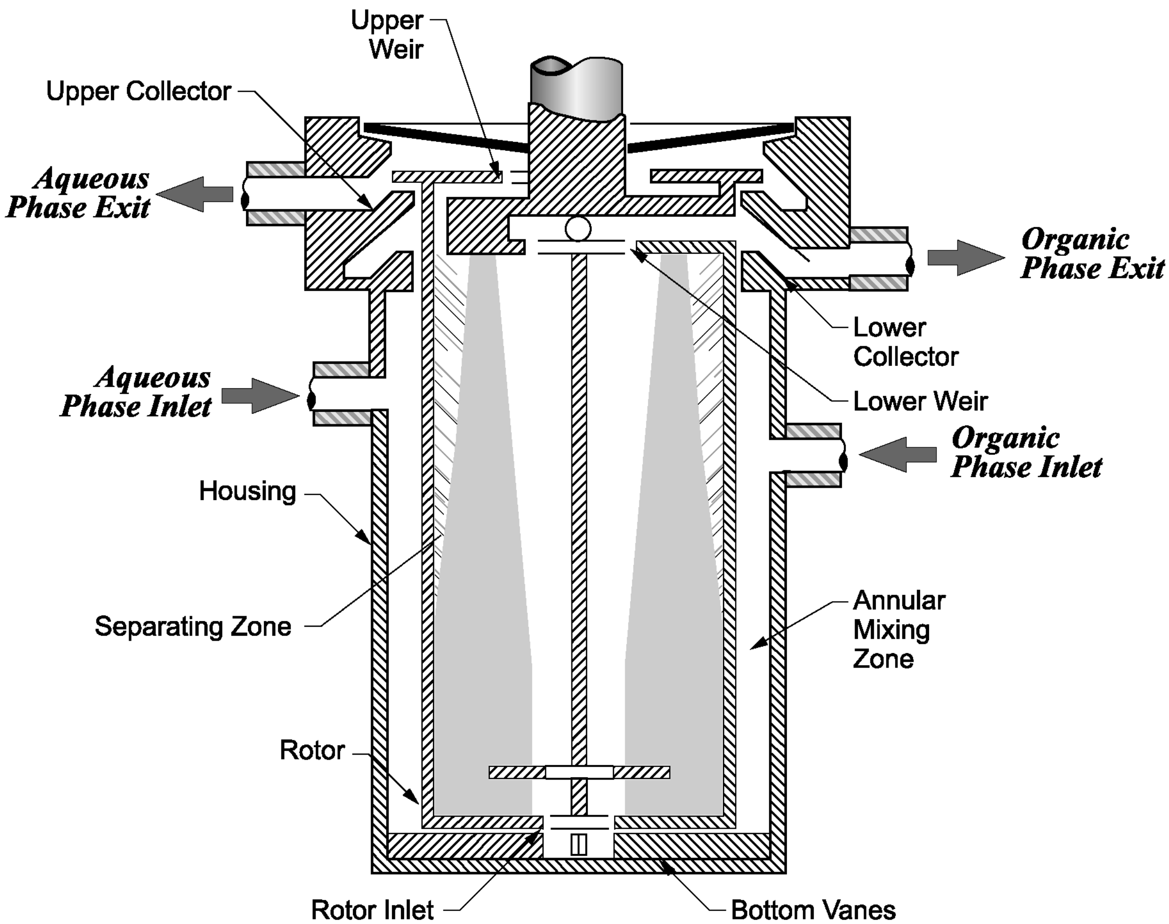
Chemical Type	Concentration in Solvent	Function
Diluent Blend of alkane hydrocarbons "Isopar® L"	Principal component	Organic phase solvent
Cesium Extractant (with complexed Cs)  Calix[4]arene-bis(*tert*-octylbenzo-crown-6) "BoBCalixC6"	0.01 M	Highly specific Cs extraction into organic phase from caustic aqueous solution
Diluent Modifier  Aromatic fluoroalcohol "Cs-7SBT"	0.5 M	Increases extractant strength for Cs, prevents precipitation and third phase formation
Suppressor $N(C_8H_{17})_3$ Trioctylamine "TOA"	0.001 M	Promotes back extraction of Cs from organic to aqueous phase during stripping operation

NW SDA EIS/Grfx/App A/Comp organic.ai

that for the Ion Exchange process, initial clarification of the salt solution in the Batch Alpha Sorption Tank would remove strontium and actinides by sorption onto monosodium titanate, followed by filtration of the monosodium titanate solids and any residual sludge, for transfer to DWPF. The separation of radioactive cesium from the salt solution by solvent extraction would take place in a multi-stage countercurrent extraction facility. The facility consists typically of an assembly of centrifugal two-phase contactors for extraction of cesium into the organic phase, scrub contactors for removing non-cesium salt constituents from the organic phase, and strip contactors for back extraction of the cesium into an acidic aqueous stream. The design and operation of the centrifugal contactors is shown in the text box on page A-19.

The cesium-containing caustic salt solution injected into the contactor assembly at the head end of the extraction section (between extraction and scrub sections) would be progressively depleted of cesium as the aqueous phase moves through the extraction contactors, and would emerge at the back end of the extraction section as a salt solution with very low cesium content. The organic phase (solvent), injected at the back end of the extraction section for countercurrent

Centrifugal Contactor Design and Operation



The separation of radioactive cesium from a high-level waste salt solution by solvent extraction utilizes countercurrent centrifugal contactors to provide high surface area interaction between the organic solvent and aqueous solution. These contactors consist of a rapidly rotating inner chamber (rotor) contained within a stationary housing, allowing mixing of organic and aqueous phases as an emulsion in the outer chamber, followed by centrifugal separation of the lesser density organic phase from the greater density aqueous phase in the rotor. The organic and aqueous phases are injected into the housing for transport through an annular mixing zone to an inlet at the bottom of the rotor. Centrifugal separation of the two phases occurs in the rotor as the emulsion flows upward, collected at the top as aqueous phase from the outer circumference and as organic phase from the center of the rotor. For extraction cycles, the cesium is transferred from the caustic aqueous phase to the organic phase and for stripping cycles it is transferred from the organic phase to an acidic aqueous phase during contactor operation.

NW SDA EIS/Grfx/App A/Centrifugal.ai

movement through the contactors, would be progressively concentrated in cesium, scrubbed to remove other salt constituents, then stripped of cesium by contact with a dilute acid aqueous stream. The strip effluent would emerge from the back end of the stripping section as a concentrated-cesium-containing aqueous solution. The organic solvent recovered from the stripping operation would be washed with dilute caustic to remove degradation products, and recycled through the extraction process, with losses made up by replacement. Occasional purging of degraded solvent during washing would generate a low-volume organic waste stream that would be stored for appropriate disposal.

Following solvent extraction separations, both decontaminated salt (raffinate) and concentrated cesium solutions (strip effluent) would be processed through stilling tanks, to float and decant entrained organic (mostly diluent) before transferring the solutions to final disposition. The decontaminated raffinate solution would be consigned to a hold tank for processing to saltstone and the strip effluent solution, assuming no concentration by evaporation, would be transferred to a hold tank for vitrification in DWPF. The wash solutions from the organic solvent cleanup would be processed to saltstone.

A.4.3.4 Direct Disposal in Grout

In the Direct Disposal (of cesium) in Grout alternative (WSRC 1998e), the HLW salt solution would be immobilized in saltstone vaults without separation of the radioactive cesium. The saltstone produced would meet acceptance criteria for near-surface disposal of low-level radioactive Class C waste (as defined in 10 CFR 61.55), but would exceed limits for Class A wastes. Treatment of the salt solution to remove strontium and actinides, as well as residual sludge, would still be required to meet restrictions on alpha-emitting radionuclides and HLW constituents in the saltstone.

If saltstone waste containing radioactive cesium was disposed in Z-Area vaults, revision of saltstone disposal procedures would be required. The existing permit issued by SCDHEC requires waste disposed in Z-Area vaults to be within Class C limits as defined in 10 CFR 61.55. SRS practice, established by DOE to minimize long-term environmental impacts, further restricts the overall average concentration of long-lived radionuclides in the Z-Area vaults at or below Class A limits. This restriction does not preclude occasional disposal of waste with higher radionuclide content if it can be shown that the waste would not produce unacceptable radiation exposure to the public, onsite workers, or inadvertent intruders. SCDHEC must be informed if the radiological content of the waste exceeds Class A limits (Martin Marietta 1992). TC

For the Direct Disposal in Grout alternative, a new facility would be constructed in Z Area, using grout production equipment modified to provide radiation shielding and enable remote operation and maintenance, because of the anticipated radioactive cesium concentrations. Direct Disposal in Grout process operations are illustrated in the flow diagram in Figure A-9. The salt solution would be collected in an H-Area tank and pumped to the Direct Disposal in Grout facility through a new Low Point Drain Tank (LPDT) facility, using the existing inter-area line. The new LPDT would be required to provide adequate shielding for the higher radioactivity in the waste stream than is present in the current feed.

In the new Direct Disposal in Grout facility, salt solution would be fed into a large Batch Alpha Sorption Tank for treatment with monosodium titanate to remove soluble radioactive contaminants other than cesium (strontium and actinides). The monosodium titanate and entrained sludge solids would be separated from the salt solution by cross-flow filtration and washed. The washed solids, collected as slurry in the Sludge Solids Receipt Tank, would be pumped through new and existing transfer lines to the DWPF melter for conversion into the glass waste form. This would be the only Direct Disposal in Grout waste stream incorporated into the DWPF waste glass production operation.

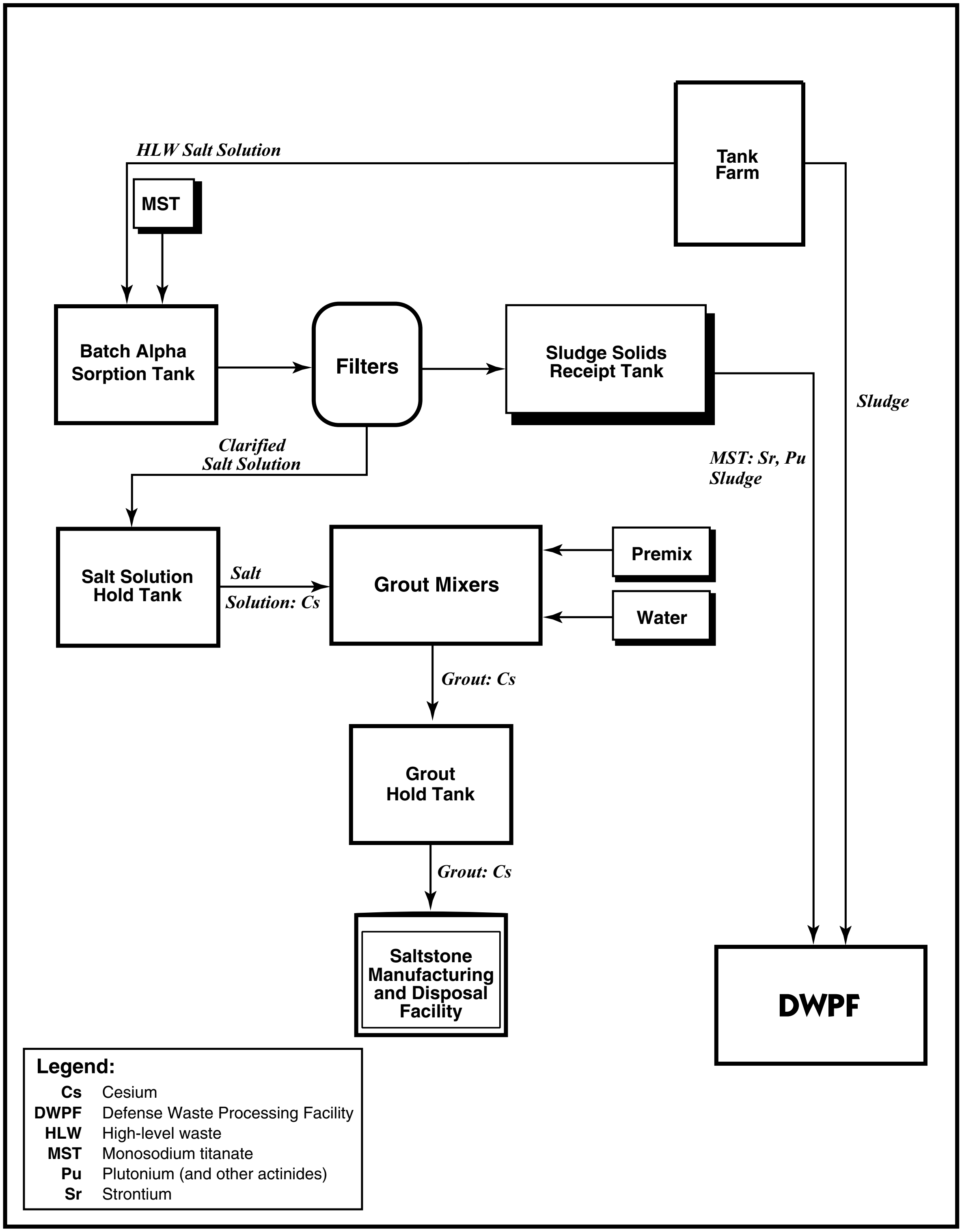


Figure A-9. Direct Disposal in Grout process flow diagram.

The clarified salt solution resulting from monosodium titanate treatment in the Direct Disposal in Grout facility would be transferred to a Salt Solution Hold Tank to be processed to saltstone.

During saltstone processing, the filtered salt solution would be pumped to a mixer and combined with flyash, cement, and slag to form a batch of grout for disposal in the saltstone vaults. The grout mixture would be pumped to a Grout Hold Tank serving as the feed tank for the Grout Feed Pumps transferring the grout to the saltstone vaults. Thirteen additional vaults would be constructed in Z Area to accommodate Direct Disposal in Grout processing. After each batch of grout was processed and transferred to a vault, the grout transfer lines, Grout Hold Tank, and Grout Feed Pumps would be flushed to remove any residual material for recycle through the process. Direct Disposal in Grout would generate no secondary waste streams.

Chemical composition of the saltstone from the Direct Disposal in Grout process is compared with that from Small Tank Precipitation, Ion Exchange, and Solvent Extraction processes in Table A-1. Expected concentrations of major radionuclides in the saltstone are shown in Table A-2. The values are from an earlier characterization of saltstone, produced for ITP processing of HLW salt solutions (Martin Marietta 1992) and adjusted for dilution by the new salt processing alternatives, based on the sodium concentrations of the saltstone feed streams.

A.4.3.5 Process Inputs and Product Streams

A general objective of the salt processing operations is the disposition of about 80 million gallons of HLW salt solution. The processing rates of the process facilities are specified to maintain a long-term average drawdown of salt solution by about 6 million gallons per year at 75 percent attainment, allowing completion of processing of reconstituted salt solution within about 13 years after facility startup. Processing within this time period is necessary to integrate the high-radioactivity salt waste components into the DWPF vitrification operations for processing with radioactive sludge components of the waste. (See key milestones textbox in Chapter 2). The throughput of all action alternatives is limited to 6 million gallons per year due to the physical constraints of removing waste from the waste tanks.

L6-23 TC

Process throughput streams for the salt processing alternatives are compared in Table A-3.

The capacity throughputs are somewhat higher than the required long-term average throughputs for Small Tank Precipitation, Ion Exchange, and Solvent Extraction facilities to allow for DWPF outages during melter changeout. The Direct Disposal in Grout facility, not closely coupled to DWPF operation, can operate at capacity throughput equal to the required long-term average throughput (6 million gallons per year).

The product outputs of the process facilities, including high-radioactivity solids slurry or solution to DWPF, processed salt solution to grout, and saltstone generated by the salt processing alternatives, are compared in Table A-4. The Solvent Extraction process would deliver a greater volume of product to DWPF than the other alternative processes because of the high volume of cesium solution (strip effluent) in the product output of that process. Salt solutions to grout and saltstone produced would be about the same for each alternative, with the ratio of saltstone volume produced to salt solution volume uniform at about 1.8.

In addition to the principal product outputs specified in Table A-4, the Small Tank Precipitation process would generate by-product benzene. About 60,000 gallons (200 metric tons) of liquid benzene would be produced annually by decomposition of the tetraphenylborate salt in the process facilities. TC

In the Small Tank Precipitation process, gaseous benzene would also be generated in the process facilities, to be dispersed into the atmosphere.

Table A-1. Chemical composition of saltstone for salt processing alternatives.

Component	Small Tank Tetraphenylborate Precipitation	Crystalline Silicotitanate Ion Exchange	Caustic Side Solvent Extraction	Direct Disposal in Grout
	Composition, weight percent [a]			
H_2O	33.70	32.88	34.03	32.57
$NaNO_3$	6.60	7.60	6.20	8.00
NaOH	1.90	2.20	1.80	2.40
$NaNO_2$	1.60	1.90	1.50	2.00
$NaAl(OH)_4$	1.20	1.40	0.94	1.40
$NaCO_3$	0.65	0.75	0.61	0.79
Na_2SO_4	0.65	0.75	0.61	0.79
$Na_2C_2O_4$	0.07	0.08	0.07	0.09
NaCl	0.05	0.05	0.05	0.06
Na_3PO_4	0.05	0.05	0.05	0.06
Na_2SiO_3	0.03	0.03	0.03	0.03
NH_4NO_3	0.03	0.04	0.03	0.04
$NaB(C_6H_5)_4$	0.03	-	-	-
Na_2CrO_4	0.02	0.02	0.02	0.02
NaF	0.02	0.03	0.02	0.03
$CaSO_4$	0.02	0.02	0.02	0.02
NaHgO (OH)	(b)	(b)	(b)	(b)
KNO_3	(b)	(b)	(b)	(b)
Salt Solution Total	46.61	47.80	45.98	48.30
Dry Blend[c]	53.39	52.20	54.02	51.70
Total	100	100	100	100

a. The values presented are taken from a previous characterization of saltstone produced during ITP processing of HLW salt solution (Martin Marietta 1992), adjusted for dilution in the new salt processing alternatives using sodium concentrations of 4.58 molar for Small Tank Precipitation, 5.31 molar for Ion Exchange, 4.30 molar for Solvent Extraction, and 5.63 molar for Direct Disposal in Grout processing, compared to 4.58 molar for ITP processing.
b. Expected present; concentration less than 0.01 weight percent.
c. Dry Blend is cement, flyash, and slag.

Issues associated with gaseous benzene generation have resulted in a number of design features that would reduce or mitigate this problem. Controlled benzene removal, because of flammability concerns, would be accomplished by operating the process vessels with a nitrogen atmosphere. The tank vent systems would be equipped with both primary and backup nitrogen purge systems (WSRC 1998e). The Ion Exchange, Solvent Extraction, and Direct Disposal in Grout processes do not have the same benzene concerns. Rather, the issue for these alternatives is radiolytic decomposition of water into hydrogen and oxygen. Air sweeps of tanks are generally considered sufficient to eliminate the danger of explosions (WSRC 1998i). However, since the consequences of an explosion are unacceptable, due to the high radioactive loading within the process tanks, the design for Ion Exchange, Solvent Extraction, and Direct Disposal in Grout facilities would include both primary and backup purge systems, comparable to those used in the Small Tank Precipitation facility.

The Solvent Extraction process would also generate a liquid organic waste requiring disposal (WSRC 2000c). The total solvent inventory for the process, consisting primarily of the diluent Isopar®L, is projected to be 1,000 gallons. This inventory is conservatively assumed to be replaced once per year. For an operational time of 13 years, the accumulated total volume of solvent requiring disposition would be 13,000 gallons.

Table A-2. Radionuclide content of saltstone for salt processing alternatives.

	Concentration (nCi/g)			
Radionuclide	Small Tank Precipitation	Ion Exchange	Solvent Extraction	Direct Disposal in Grout
Technetium-99	33	38	31	40
Ruthenium-106+d[a]	17	20	16	21
Cesium-137+d[a]	10	12	9	254,000[b]
Tritium	10	12	9	12
Antimony-125	3.3	3.8	3.1	4.0
Promethium-147	2.0	2.3	1.9	2.4
Samarium-151	1.0	1.2	0.95	1.2
Strontium-90+d[a]	0.35	0.40	0.33	0.42
Europium-154	0.33	0.38	0.31	0.40
Selenium-79	0.16	0.19	0.15	0.20
Europium-155	0.16	0.19	0.15	0.20
Cobalt-60	0.11	0.12	0.10	0.13
Tellurium-125m	0.10	0.12	0.09	0.12
Tin-126+d[a]	0.07	0.08	0.07	0.08
Cesium-134	0.03	0.04	0.03	440
Tin-121m	0.01	0.02	0.01	0.02
Iodine-129	0.01	0.01	0.01	0.01
Nickel-63	0.01	0.01	0.01	0.01
Antimony-126	0.01	0.01	0.01	0.01
Carbon-14	0.003	0.004	0.003	0.004
Cesium-135	0.00002	0.00002	0.00003	0.26
Other beta gamma	3.3	3.8	3.1	4.0
Plutonium-238	0.03	0.03	0.03	0.03
Plutonium-241	0.02	0.02	0.02	0.02
Americium-241	0.07	0.08	0.07	0.08

nCi/g = nanocuries per gram.
a. +d = with daughter product.
b. Cesium-137+d content of the saltstone for Direct Disposal in Grout alternative corresponds to 225 Ci/m^3 of cesium-137 (WSRC 1998a,k).

Table A-3. Salt solution processed.

Alternative	Required processing rate (million gallons)[a] per year	Long-term average throughput capacity (million gallons per year)[a]	Throughput limitation
Small Tank Precipitation	6.9	6.0	Salt removal rate from waste tanks
Ion Exchange	6.9	6.0	Salt removal rate from waste tanks
Solvent Extraction	6.9	6.0	Salt removal rate from waste tanks
Direct Disposal in Grout	6.0	6.0	Salt removal rate from waste tanks

L6-23

Source: WSRC (1998e).
a. The required processing rate for the salt processing facilities exceeds the long-term average to allow for downtime when DWPF is not operating, except for the Direct Disposal in Grout facility which can operate at the required salt removal rate even when DWPF is not operating. L6-23